Case study: Distillation column control for separating chemical petrol in a refinery

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This case study deals with control of distillation which is used for separating of two fluidic components from a liquid mixture based on their different boiling points. For example, ethylalcohol and water can be separated from wine, as the boiling point of the alcohol is 78°C and that of the water is 100°C at atmospheric pressure. The same technology is used for separating chemical petrol in refineries. As a distillation column is a TITO (Two-Input, Two-Output) process (a special case of a multivariable process), two important control techniques can be applied:

1. Suitable pairing of the two manipulated variables to the two controlled variables if only two separate controllers are used.
2. Designing a TITO controller to decouple the control of the two control circuits, i.e. the first manipulated variable influences only one of the controlled variables, and the other manipulated variable affects only the second controlled variable.

The first task will be performed by the RGA (Relative Gain Array) method and the second task by model based predictive control.

Prior to introducing the control task the distillation process is explained briefly. For simplicity a binary mixture with only two components is dealt with. Distillation is the most common technic to separate liquids from each other. It is applied especially in the chemical and petrochemical industry.

1. Batch distillation

Distillation is used to separate liquids with different boiling points from each other. If the mixture is heated and the temperature achieves the boiling point of the component with lower boiling point then this component vaporizes and leaves the mixture. The vapor is cooled in the condensator and gets liquid phase again.

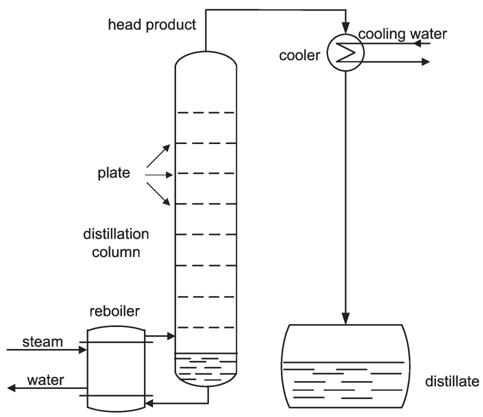

Fig.1: Batch distillation

Through this process the not vaporized fluid is separated from the vaporized fluid. However the concentrations of both separated fluids change permanently. The process is called batch distillation as no mass transfer into or out of the system occurs, that means no external feed is supplied and the separated liquids are not lead away. A further problem is that the boiling points depend also on the pressure, thus both the temperature and the pressure have to be controlled. The scheme of batch distillation is shown in Fig. 1.

2. Continuous distillation

With continuous distillation the educt is fed continuously into the process and the two products with different concentrations are lead away also continuously. The advantage of continuous distillation over batch distillation is that the concentrations and quantities can be kept constant, so the process is in a steady state.

The aperture that facilitates continuous distillation is called column (Fig. 2.).

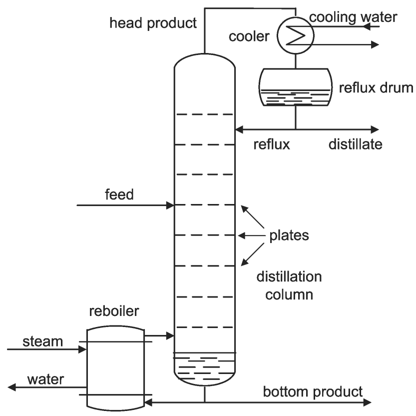

Fig.2: Continuous distillation

The educt is fed into the middle of the column. The not vaporized liquid is streaming downwards and leaves only partly the column as bottom product. A part of the liquid is vaporized in the reboiler and the steam streams upwards. At the top of the column the steam leaves the column and is condensed by cooling to liquid. This liquid is divided into two streams. The distillate leaves the column as the product with the lower boiling point. This is called head product. The other part of the liquid is fed back into the column as reflux. The downwards streaming fluid meets the upwards streaming vapor at several plates where the steam is condensed and the fluid is vaporized until vapor and fluid are in thermal and concentration balance. At a given plate (that means at a certain height) the concentration is the same as the concentration of the feed. This is the point where the educt has to be fed into the column not to disturb the operation.

3. Instrumentation and control of the distillation column

There are several variables, which can be affected by control. Nearly all of them affect each other as well.

- Very often the feed is the product of another distillation column and cannot be influenced by the actual unit. The temperature (TT-81) can be manipulated easily by pre-heating heat exchanger.

- At the bottom of the column, the level (LT-31) has to be controlled to prevent the reboiler from running dry and the column from flooding. In addition, quality (QT-31) of the bottom product has to be controlled. To perform this, the amount of the heating steam and the amount of product can be manipulated by the corresponding valves.
- At the top of the column, the liquid level of the vessel (LT-21) has to be controlled. In addition, quality of the head product (QT-41) has to be controlled. The same has to occur with pressure (PT-11) in the column. To perform this, the flow of reflux and the head product stream can be regulated by the corresponding valves. In addition, the amount of cooling of the condenser can be regulated to control the pressure. Fig. 3 shows water cooling, in some applications natural air cooling is used. The air cooling is less expensive but cannot work effectively on hot days.

Instead of the qualities of the products, the temperatures near to the ends of the column (TT-61 and TT-71) can be controlled. This is a cheaper solution, but it has to be taken into account that the concentration is affected not only by the temperature, but also by the pressure.

The instrumentation of a distillation column is shown in Fig. 3.

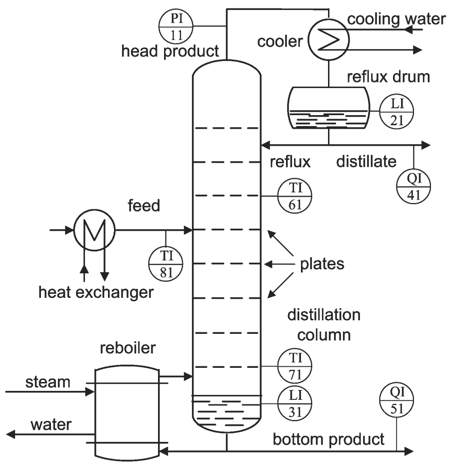

Fig. 3: Distillation column with measurement instrumentation (.T denotes measurement transmitter)

Some input variables are not constant and may disturb the operation of the control. The temperatures and pressures of the cooling and heating media may change. The feed is restrained by the feeder system so its parameters (flow rate, temperature and composition) may change. The aim of the control is to compensate the effect of all these disturbances and to ensure the desired concentrations of both (distillate and bottom) products.

4. Control structures when only one quality is controlled

There are several possible strategies to control only one product quality in a distillation column (e.g. Buckley et al., 1985). Experiences show that

- The compositions of the product streams are more sensitive to a change in the mass balance as to a change in the energy balance.
- Big dead times should be avoided, e.g. bottom liquid flow should not controlled by reflux flow, etc.

There are some well established regulation methods, as e.g. controlling the pressure by cooling. The most common four combinations for controlling only one product quality are shown in Table 1.

Table 1: Different control strategies for distillate quality control

Case	Manipulating variable to control vessel level	Manipulating variable to control bottom level	Manipulating variable to control the quality (composition)	Not used manipulating variable
1	Reflux flow	Bottom product flow	Distillate flow	Heating stream flow
2	Distillate flow	Heating stream flow	Bottom product flow	Reflux flow
3	Distillate flow	Bottom product flow	Heating stream flow	Reflux flow
4	Distillate flow	Bottom product flow	Reflux flow	Heating stream flow

The most typical control of distillation concentration is shown in Fig. 4. The vessel level is controlled by the reflux flow. The bottom liquid level is controlled by the bottom product flow. The reboiler steam flow is free and can be used to optimize cost efficiency.

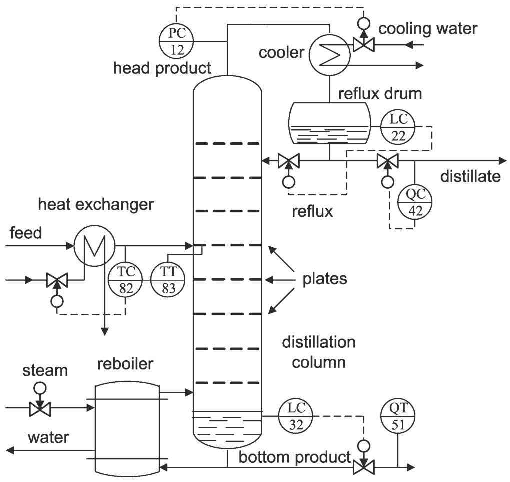

Fig. 4: Distillation column with controlling the distillate concentration

5. Control structures when both qualities are controlled

There are some possible control combinations as:

Case1: energy balance control structure:
- distillate concentration by reflux flow,
- bottom product concentration by heating steam flow,
- condensate vessel level by distillate flow,
- bottom liquid level through the bottom flow

Case2: mass balance control structure:
- distillate concentration by distillate flow,
- bottom product concentration by heating steam flow,
- condensate vessel level by reflux flow,
- bottom liquid level through the bottom flow

Case 3: mass balance control structure:

• distillate concentration by reflux flow,
• bottom product concentration by the bottom stream,
• condensate vessel level by distillate flow,
• bottom liquid level by heating steam flow

Fig. 5 shows the energy balance control structure, which is a common strategy. It can be seen, that both manipulated variables influence both controlled variables. Increasing the reflux flow cools not only the top product but also the bottom product. Increasing the heating steam flow increases not only the bottom product temperature but also the temperature of the top product. The control is relative fast but coupled. It is the task of controller design to try to decouple this multivariable (in this case TITO) control. That means changing the set value of only the distillate or only the bottom product concentration should not cause any change in the other quality.

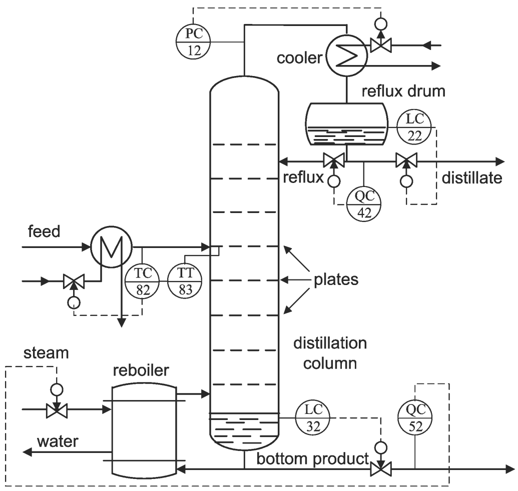

Fig. 5: Distillation column with controlling the distillate concentration

A better solution for control is to perform a ratio control which is not treated here.
Fig. 6 shows an input/output model of the distillation column for quality control.

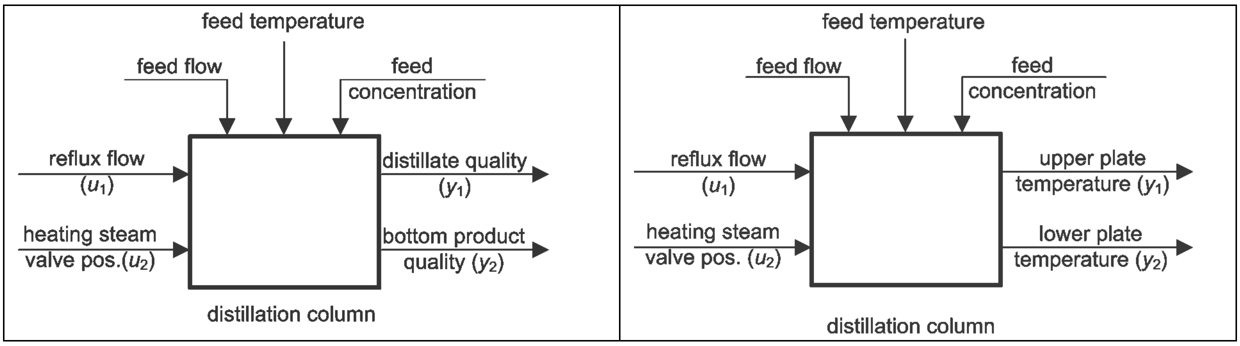

Fig. 6.: Input/output model of the column for quality control (left side) and for the related temperature control (right side)

The two *controlled variables* are:

- distillate quality (y_1)
- bottom product quality (y_2)

The *manipulated variables* are:

- reflux flow (u_1)
- heating power (u_2)

The main *disturbances* are:

- feed quantity (z_1)
- feed temperature (z_2)
- composition of the feed (z_3)

For quality (composition) measurement a chromatograph can be used which is expensive equipment and the measurement time is about 15 minutes. Instead it is usual to control the temperature of the lower and the upper plates (or the plates close to them), as the temperature of the plates is related to the composition, and temperature measurement is cheap and the measured signal is immediately available. As the concentration depends also on the pressure, generally the Pressure Compensated Temperature (PCT) is controlled. If also a chromatograph or results of laboratory analysis are available, cascade control can be applied. In the sequel the model of a simple distillation column is investigated where the controlled variables are the PCT-s of the upper and lower plates, which are characteristic for the concentrations of the distillate and the bottom product, respectively. The process model is given in Table 2. The sub-processes are modeled by first-order plus dead time models with gain K_p, dead time T_d and time constant T.

Table 2: Input/output model of the distillation column

Manipulated variable	Controlled signals	
	Top (distillate) PCT (y_1)	Bottom PCT (y_2)
Reflux (u_1)	T_{d11} = 3 min K_{p11} = 0,0275°C/(t/d) T_{11} = 9,4 min	T_{d21} = 10 min K_{p11} = 0,0201°C/(t/d) T_{21} = 12,6 min
Heating power (u_2)	T_{d12} = 10 min K_{p12} = 3,32°C/MW T_{12} = 8,3 min	T_{d22} = 3 min K_{p22} = 4,45°C/MW T_{22} = 6,5 min

Fig. 7. shows the process model with the transfer functions. (In the Laplace transform forms of the transfer functions the time constants are given in minutes.)

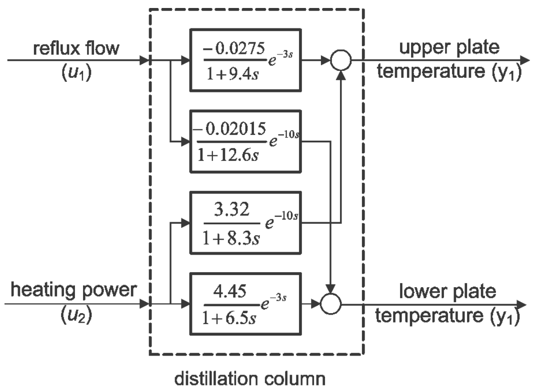

Fig. 7: Input/output process model

6. **Demonstrating the coupling effect in distillation column control**

Traditionally the distillate and the bottom product quality were controlled independently. Fig. 8 shows this case and also gives the structure of a decoupling controller.

The dashed blocks indicate the TITO process to be controlled and the controller, respectively. The process is split into four sub-processes. The first number in the indices denotes the target of the process, in this case the quality. The second number indicates the source of the process, in this case the manipulated variables.

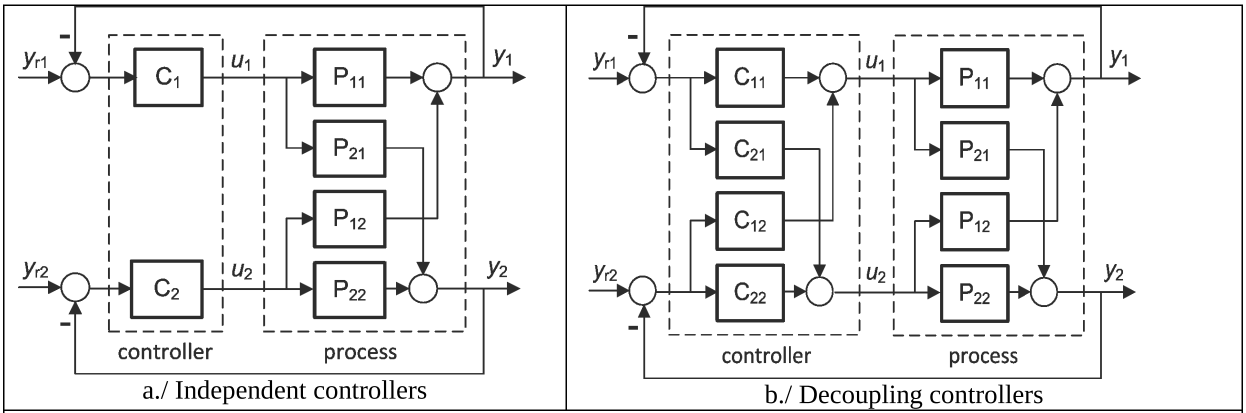

Fig. 8: Independent (left side) and decoupling (right side) controllers with the TITO process

In the sequel the difference between the independent and the decoupling controls is shown based on a simplified model of a distillation column.

Fig. 9 presents the pressure controlled temperatures of the top and the bottom plates in the cases of the independent and the decoupling control. The control algorithm realized predictive control optimizing a cost function (which will be presented later). Fig. 9a shows that with the independent control the quality which is not controlled is altering if the set value of the other quality was changed. That means that the bottom product quality will not remain constant if only the distillate quality set value is changed, and vice versa. This coupling effect could be almost perfectly eliminated by the decoupling controller see Fig. 9b.

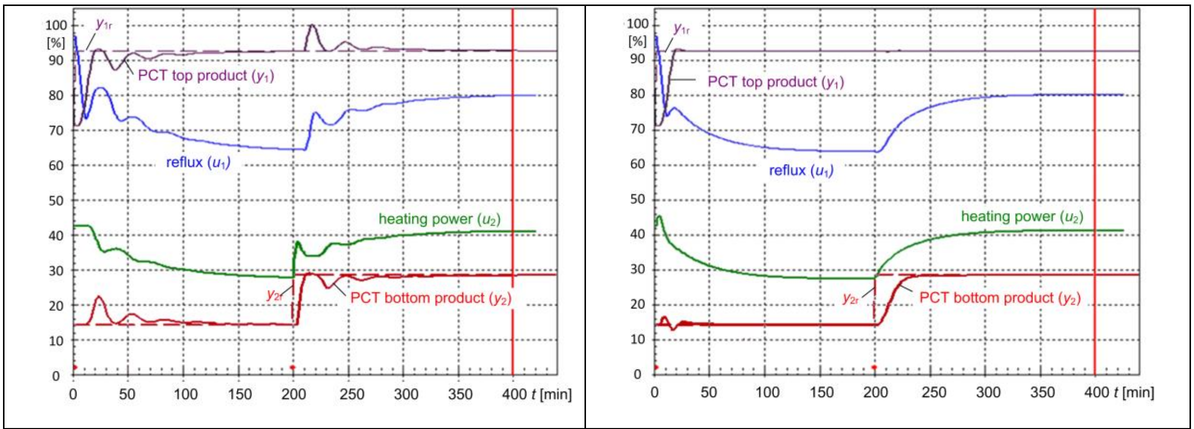

Fig. 9: Independent (left side) and decoupling (right side) control of the pressure controlled temperatures of the top and bottom plates

In all simulations the variables were scaled in relative units from 0% to 100%. The range of TOP-PCT is between 50°C and 64°C, BOT-PCT moves between148°C and 162°C, reflux flow changes between 200 and 520 t/d, and the range of the heating power is between 3 MW and 10 MW. Initially the value of the TOP-PCT was 60°C and that of the BOT-PCT was 150°C. In t=0 min the set value of TOP-PCT was increased from 60°C to 63°C and in t= 200 min the set value of BOT-PCT was increased from 150°C to 152°C.

Why is a decoupling controller better than the two independent controllers? The reason is that the decoupling controller compensates the coupling effect in the process. Let us consider for example the case when the set value of the distillate concentration is increased, which means decrease of the temperature set point. The independent distillate controller C_1 decreases the reflux flow, and this decreases the temperature of both the distillate and the bottom products. It takes a long time until the controller C_2 of the bottom product quality compensates this disturbance. If a well-tuned TITO controller is applied then the controller C_{21} decreases the heating power of the reboiler to compensate the effect of the reflux reduction. (Let us remark that controller C_{11} of the TITO controller is not equal to the independent controller C_1.)

7. Analytical determination of the intensity of the coupling effect

The most direct way to control the qualities is to use two independent controllers as shown in Fig. 8a. The two control loops disturb each other via the coupling terms in the process model. The coupling can be reduced by a proper pairing of the two manipulated and controlled signals. This can be done by using the so called RGA (Relative Gain Array) method (Bristol, 1966).

First arbitrary pairing between the manipulated and control variables is supposed. (For example the distillate quality is controlled by the reflux flow and the bottom product quality is controlled by the heating power.) Then the static gain of process P_{11} has to be determined in two different cases.

1. The control of sub-process 2 is in manual mode, (i.e. the manipulated variable u_2 remains unchanged, Δu_2=0) (Fig. 10a).
2. The control of sub-process 2 is in automatic mode, i.e. the controlled signal y_2 remains constant, Δy_2=0) (Fig. 10b)

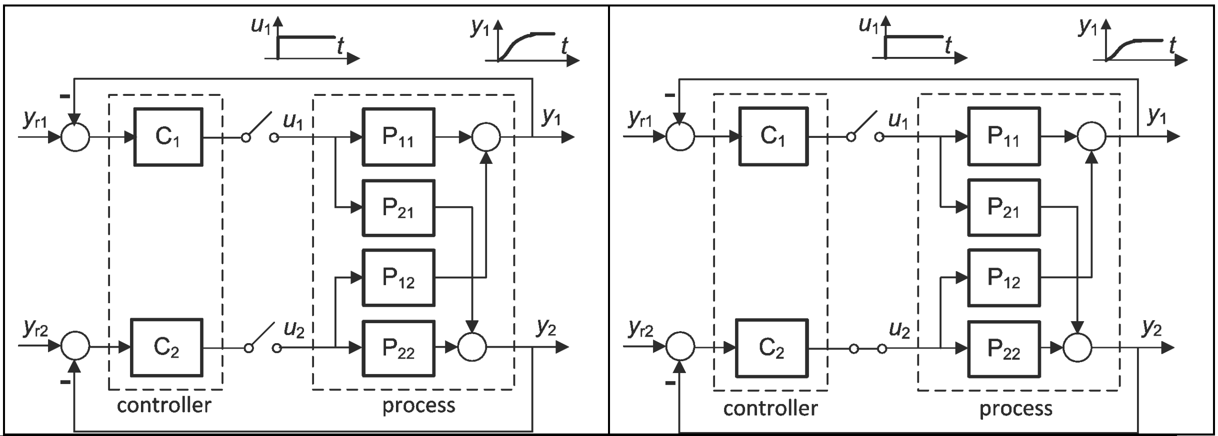

Fig. 10: Static gain $K_{p11} = \frac{\Delta y_1}{\Delta u_{11}}$ with control loop 2 in hand (left side) and automatic (right side) mode

If the ratio $\lambda_{11} = \frac{K_{p11,\text{not\_controlling\_loop2}}}{K_{p11,\text{controlling\_loop2}}}$ is 1, then there is no coupling effect and the two control loops can be tuned independently. In any other case a decoupling controller is the right choice. Using, however, two independent controllers it should be taken into account that both controllers can handle both cases, that means the other controller works in automatic mode (regular case) or the other controller is in hand mode (e.g. because of a sensor fault).

Fig. 11 shows the results of the two simulation experiments. The ranges of the variables are scaled as in Fig. 9. Initially the TOP-PCT was 60°C and the BOT-PCT was 150°C, the corresponding manipulated variables values in steady state were as follows: reflux flow 510,4t/day and heating power 42 MW. At t=0 min the reflux flow was decreased from 510,4t/day to 440t/day and it was investigated how the process reaches its new steady-state. Fig. 11 shows the two cases when the BOT-PCT was not controlled and when it was controlled.

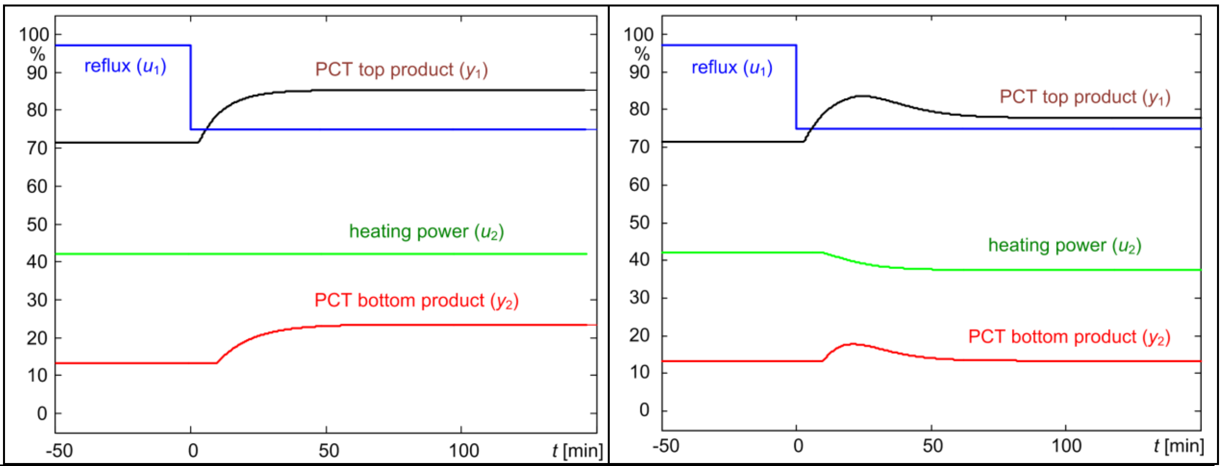

Fig. 11: Step responses to reflux flow change without controlling of the bottom product quality (left side) and with controlling it (right side)

The relative gain can be calculated as

$$\lambda_{11} = \frac{K_{p11,\text{not\_controlling\_loop2}}}{K_{p11,\text{controlling\_loop2}}} = \frac{\dfrac{61,9406°\text{C} - 60,0044°\text{C}}{440\text{t/d} - 510,4\text{t/d}}}{\dfrac{60,8864°\text{C} - 60,0058°\text{C}}{440\text{t/d} - 510,4\text{t/d}}} = 2.2$$

Value 2.2 shows that the gain is higher if the bottom product quality is not controlled, e.g. in case of a sensor fault. In this case a well-tuned control loop (for the regular case if both controllers work) may oscillate.

There is another possibility to calculate the relative gain. It can be calculated from the static gains of the sub-models of the process.

$$\lambda_{11} = \frac{K_{p11,\text{not\_controlling\_loop2}}}{K_{p11,\text{controlling\_loop2}}} = \frac{1}{1 - \dfrac{3.32°\text{C/MW} \cdot (-0.0201°\text{C/(t/d)})}{(-0.0275°\text{C/(t/d)}) \cdot 4.45°\text{C/MW}}} = 2.2$$

The next task would be to check whether the other pairing of the manipulated and controlled variables, that means controlling the distillate quality by the heating power and the bottom product quality by the reflux flow, would be more suitable. According to the RGA method the alternative pairing results in relative gain

$$\lambda_{11,\text{alternative}} = 1 - \lambda_{11} = 1 - 2.2 = -1.2$$

This control structure is not preferred because of the static gain changes sign depending on whether the second controller is in automatic or in manual mode. This means that a sensor fault may lead to unstable control.

Based on the above considerations the decoupling control is prior to the two independent controllers. Instead of designing a TITO decoupling controller a model based predictive controller is designed which possesses a decoupling feature as well.

8. Model based multivariable predictive control

A TITO decoupling controller consists of four sub-controllers, two of them really controlling and two compensating the disturbing coupling effects in the process. This linear controller does not consider the physical constraints. Limitation in the manipulated variable is can be taken into account, but constraints in the controlled and in other calculated variables cannot be handled in this structure. Therefore another control technology, model predictive control will be used in the sequel for distillation column control.

The basic idea of the method is to consider the output variables not only in the actual time point but also during their course in a future horizon. This goal is achieved first by heuristic choice of the manipulated variable sequence and simulation of the future course of the process variables. If the future course of the controlled and the constrained variables is not satisfactory, then new manipulated variable sequences are tried out until the control behavior becomes satisfactory, that means the target and the constraints are fulfilled.

Predictive control is widely used in petrochemical plants, as distillation is a relatively slow process allowing multiple simulation of the future course of the process outputs considering also the fulfillment of the constraints. But generally the following quadratic cost function is minimized (e.g., Camacho and Bordons, 1998, 2004):

$$J = \sum_{i=1}^{N_y} \sum_{n_{ei}=n_{e1i}}^{n_{e2i}} \frac{\lambda_{yi}}{\lambda_{yi}^*} [y_{ri}(k+d_i+n_{ei}) - \hat{y}_i(k+d_i+n_{ei} \mid k)]^2 + \sum_{i=1}^{N_u} \sum_{j=1}^{n_{ui}} \frac{\lambda_{ui}}{\lambda_{ui}^*} \Delta u_i^2(k+j-1) \Rightarrow \underset{\Delta u_i}{MIN}$$

where the meaning of the variables is the following:

- $\Delta u_i(k)$: the i-th manipulated variable,
- $\hat{y}_i(k+d_i+n_{ei} \mid k)$: the i-th predicted controlled signal,
- $y_{ri}(k+d_i+n_{ei}|k)$: the i-th reference signal given $n_{ei}+d$ steps ahead is, with
- d_i: the discrete-time delay of the i-th controlled signal and
- n_e: prediction step over the dead time.

The predicted control signals should include the main and auxiliary controlled signals and the soft constrained variables, as well.
The tuning parameters of the controller are as follows:

- N_y: number of the controlled signals,
- N_u: number of the manipulated variables,
- n_{e2}-n_{e1}+1: length of the optimization (prediction) horizon,
- n_u: length of the control horizon,
- λ_{yi}^*: scaling factor of the i-th controlled or constrained signal,
- λ_{ui}^*: scaling factor of the i-th manipulated variable,
- λ_{yi}: weighting factors of the i-th controlled or constrained signals,
- λ_{ui}: weighting factors of the i-th manipulated variable.

In practice often a set range is given instead of a set-value. The cost function is zero if all the controlled variables are within their ranges and there is no change in the manipulated variables.

This predictive control algorithm was applied successfully for control of an industrial distillation column Volk et al., 2005; Haber et al., 2011).

9. Description of a distillation column for separating chemical petrol

Fig. 12a shows the distillation column to be controlled and Fig. 12b gives its piping and instrumentation diagram.

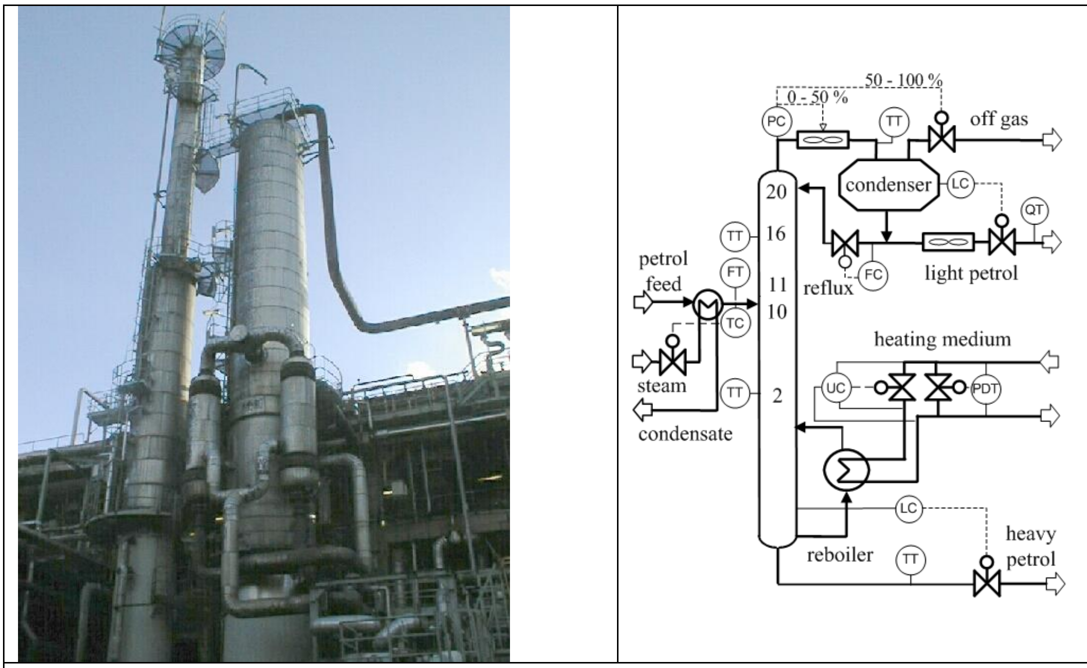

Fig. 12: Distillation column: photo (on the left) and piping and instrumentation diagram with basic control loops (on the right)

The feed is chemical petrol from the desulfurization. The top product is light petrol with boiling point between 30 °C and 65 °C. The bottom product is heavy petrol with boiling point between 65 °C and 180 °C. The reflux can be manipulated by a PI flow control loop. The pressure is controlled by changing the pitch of the fins of the air coolers. The level controls are performed by the corresponding product streams. The reboiler power is controlled by the flow of the heating oil.

10. Controlled and auxiliary controlled variables, manipulated variables and disturbances

The control targets are to keep both product qualities within a set range under consideration of constraints and to minimize operation costs.

The main variables which have to be kept constant or have to be minimized are as follows:

1. *Distillate quality:*
   The concentration of the heavy petrol in the distillate has to be kept below 2,5 %. The concentration is considered with FBP (Final Boiling Point), i.e. the boiling temperature of the product if 99 % of the product has been evaporated.

2. *Bottom product temperature:*
   Instead of the concentration of the bottom product the pressure compensated temperature of the bottom product is controlled. This variable is related to concentration and its measurement is much cheaper.
3. *Furnace load:*
   The column temperature is controlled by changing the hot oil flow through the reboiler. In order to keep the oven operation smooth, the heating medium flow is kept constant by using a bypass. Therefore the pressure difference is measured by the transmitter PDT in Fig. 12b.
4. *Column pressure:*
   The column pressure has to be kept at its minimum value in order to reduce the energy consumption (reboiler heating).

So the main controlled variables are:
1. The distillate quality, measured by its FBP
2. The pressure compensated temperature of the bottom product
3. The heating power
4. The column pressure which is to be minimized

There are two auxiliary controlled variables which contribute to a better operation:
1. *Feed temperature:*
   The feed temperature should be equal to the feed tray temperature not to disturb the operation of the column.
2. *Pressure compensated temperature of top plate:*
   The PCT of tray 16 is calculated and controlled by an inner cascade loop. The top PCT shows earlier a distillate quality change than the measured FBP.

11. Constrained signals

Limitations (constraints) in the values and rates of the signals have to be considered already in the design phase otherwise simulation (without constraints) and the real behavior differ. The real control becomes slower than planned and quality requirements cannot be fulfilled.

Constraints can be divided into two types:
1. *Hard constraint:*
   Hard constraints cannot be violated, for example, the maximum flow in a pipe or valve.
2. *Soft constraint:*
   Soft constraints should not be violated but can be violated for a short period, for example, the allowed environmental pollution. If this limit is exceeded occasionally a fine has to be paid but the operation does not have to be stopped immediately.

Hard constraints are used as limitations during the numerical optimization of the target (cost) function. Soft constraints can be handled by additional quadratic terms in the quadratic cost function.

The following limits should be considered for the manipulated variables:
1. *Set value of the reflux basic controller:*
   Minimum, maximum and changing rate
2. *Set value of the pressure basic controller:*
   Minimum, maximum and changing rate

3. *Valve position of the feed temperature controller:*
   Minimum and maximum
4. *Valve position of the temperature control of the heating medium:*
   Minimum and maximum

The following soft constraints should be considered:
1. *Valve position of the pressure controller:*
   Maximum constraint
2. *Heating medium:*
   Minimum and maximum
3. *Position of the bypass valve at the heating medium pipe:*
   Minimum and maximum
4. *Set value of the reflux basic controller:*
   Minimum and maximum
5. *Violation of the furnace duty:*
   Maximum constraint: in order to protect the furnace against overheating.

12. Process and disturbance model

Any controller design requires a process model. The process model and also the disturbance model have to be determined.

According to our experience does not keep its steady state for a longer time, because of the influences caused by the operation of neighboring units or raw material fluctuations. Therefore, instead of a pseudo-random signal some step tests were applied for experimental identification. Fig. 13 shows such a test sequence which ensures that at the end of the test the working point will not change.

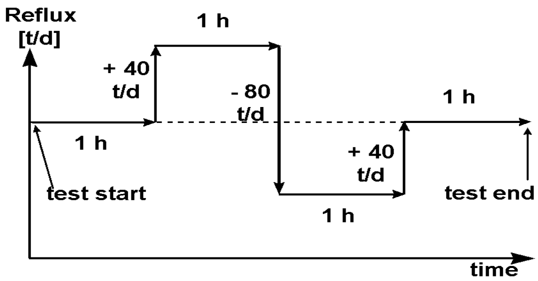

Fig. 13: Step test (e.g. for the reflux flow)

From the measured records the parameters of first-order models including dead time were identified by least squares parameter estimation algorithm. These step tests took the most time (some weeks) of the controller design procedure, as often other variables also have changed disturbing the measurement.

Fig. 14 shows the process model (left side) and the disturbance model (right side) of the column obtained by the experimental identification. The model structure is shown, and the numerical values of the static gains (G), time constants (T_1) and dead times (T_d) are also given (the time constants are given in minutes). The identification results are analyzed in details in references Volk et al., 2005; Haber et al. 2011. It has to be mentioned that second order models did not give significantly better results than the first-order models.

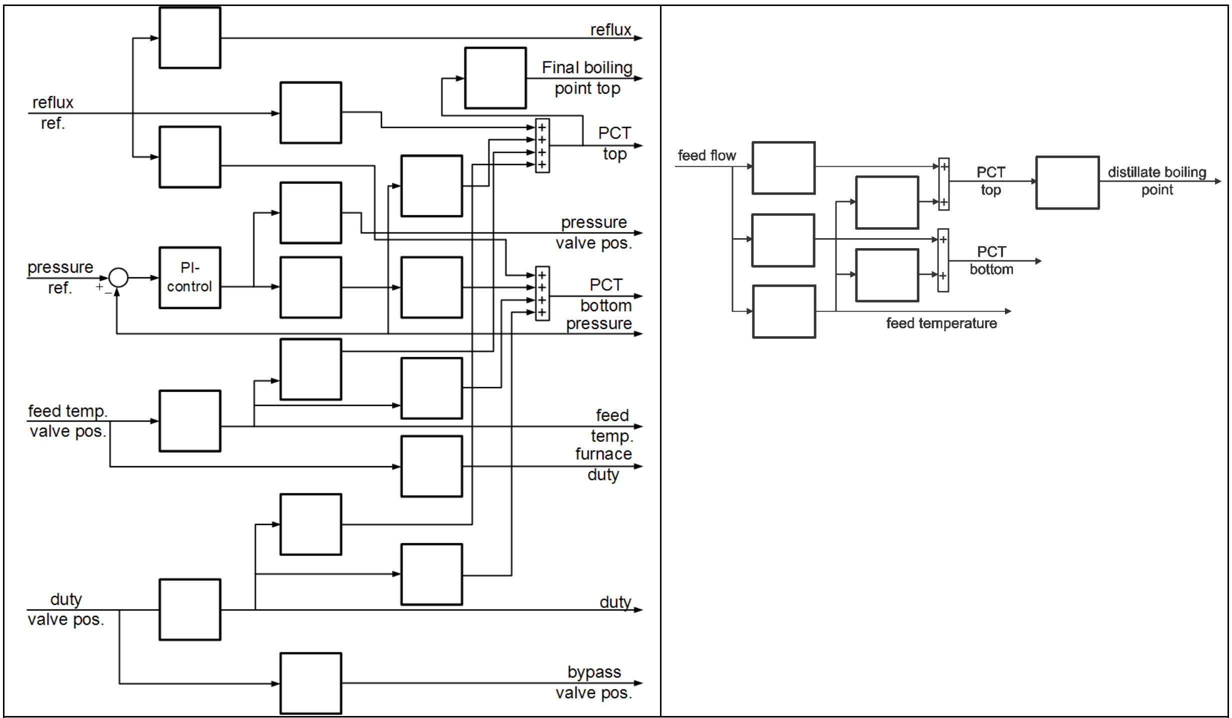

Fig. 14: Process (on the left) and measured disturbance (on the right) models

The process model differs a bit from the usual input/output models. The basic reflux control loop, that means the relation between the reflux reference value and the reflux flow is modelled by an overall transfer function (with static gain approximately 1 and small time constant), while the pressure control loop is modelled with its PI controller and the open-loop pressure model (between the valve position and the column pressure). This modeling has the advantage that the manipulated variable of the pressure control loop, i.e. the valve position can be modelled and used as a soft constrained variable. Further on any disturbance influencing the pressure can be quickly compensated by the PI controller and the predictive controller of the bottom and top products does not have to change the corresponding manipulated signals.

Typical disturbances of distillation column are changes in the feed parameters, like temperature, composition and flow. Here only the flow was considered as a measured disturbance.

13. Control simulation and controller tuning

Fig. 15 shows the simulation of the control, when a stepwise feed flow decrease (disturbance) occurs at time point 30 min without optimizing and with optimizing the weighting factors in the cost function. (Initially all the weighting factors were chosen equal to 1.) As it is seen, after having optimized the weighting factors (by repeated simulations), the control became faster. Also the pressure decreases as the optimizing controller minimizes the pressure reference value while fulfilling the quality requirements.

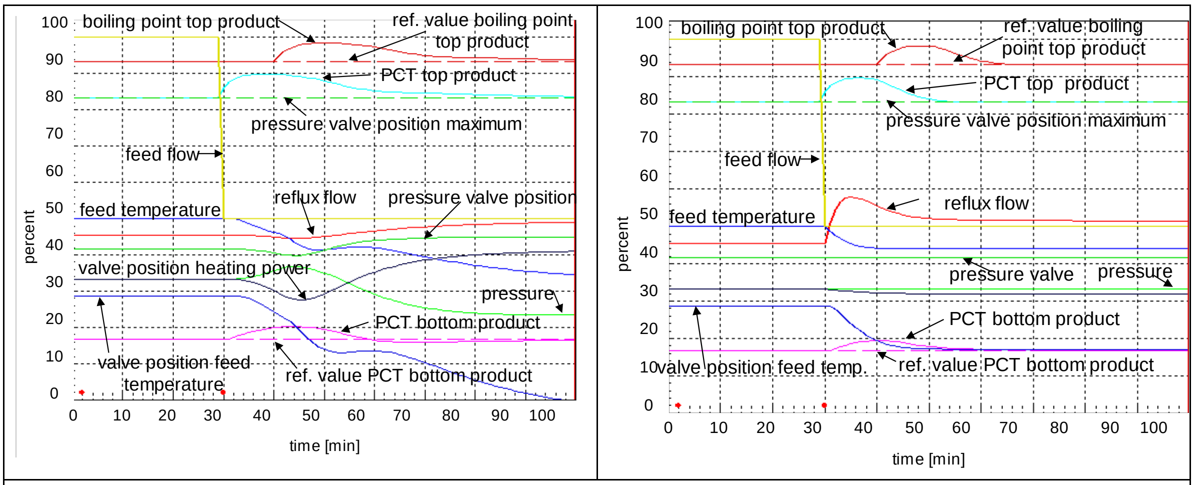

Fig. 15: Behavior of the control in case of a feed step without (left) and with (right) optimizing the weighting factors

14. Real-time control

Figs. 16a and 16b show the real-time control during the first 10 days of implementation of the predictive control. In the first 6.5 days only the basic control loops worked and then (marked by a dashed vertical line) the predictive control was started and the first 3.5 days of its behavior is shown.

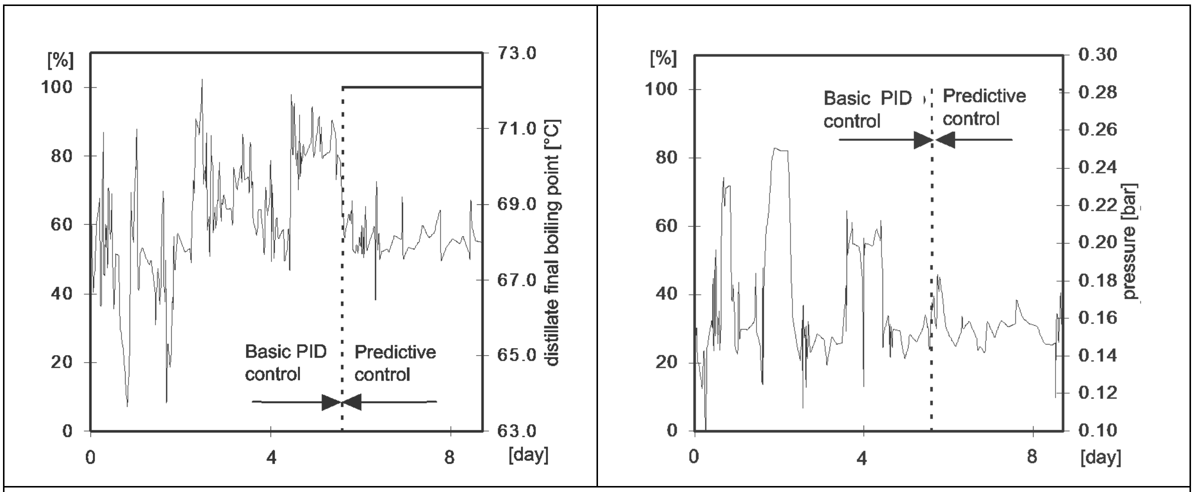

Fig. 16: Control of final boiling point of the top product (left) and control minimizing the column pressure (right)

The main controlled variable, the final boiling point of the top product is kept almost constant, and the reduction of the peak to peak value due to introduction of predictive control is about 3:1. Also the column pressure control became much smoother (reduction of the peak to peak values is better than 4:1) and the pressure keeps its minimum possible value.

15. Summary

Distillation is a basic technology in refineries. It was shown for a column separating chemical petrol in a refinery that different control strategies can be used if only one concentration (usually the distillate quality or if both concentrations (top and bottom product qualities) are controlled. In the first case a SISO control is applied. In the second case a coupled TITO process has to be controlled and the controller design is complicated. Paring of the two manipulated variables (heating and reflux flow) to the two concentrations to be controlled is not a trivial task. The RGA method was used to select the better control structure. Even having the better pairing the two controllers cannot be designed independently, iterative design is required. Using only two controllers does not compensate the coupling effect in the column, which means that changing of the heating or the reflux flow influences simultaneously both concentrations. A decoupling controller containing four controller terms can optimally control the TITO process. As tuning of such a complex controller is a difficult task, predictive control strategy was applied for minimizing a quadratic control cost function.

The controller keeps the distillate and bottom concentration constant while fulfilling different practical physical constraints. Furthermore, the controller is able to compensate measured disturbances, e.g., feed flow changes, by disturbance feed-forward control. Applying predictive cascade control by measuring an inner variable (top PCT) the main control variable (final boiling point of top product) can be controlled much faster.

The weighting factors of the control errors and of the manipulated variable increments were roughly tuned based on physical considerations and fine-tuned in several simulation runs in an iterative way. Some plots show the controlled and manipulated variables before and after the implementation of predictive control. It is demonstrated that the quality fluctuation became much smoother. Furthermore, using a lower pressure reference value led to less duty what means energy saving. By keeping the pressure difference on the bypass valve constant also, the loading flow of the coupled furnace was kept constant, i.e. the operation of the furnace became smoother.

Through decreasing the variance of the final boiling point of the top product the reference value could be increased and heavier (cheaper) components in the expensive top product could be sold. This is shown by the following calculation assuming that the feed flow is 3136 t/day. Table 3 shows the mean value and the standard deviation of the final boiling point of the top product with the PI and with the predictive control. Because of the reduction in the standard deviation of the controlled variable the reference value can be increased. Thus more bottom and less top product can be produced as shown in Table 4. (That means that more heavy components can be sold in the top product.)

Table 3: Distillate quality with different control techniques

Control	Final boiling point	
	Mean value [°C]	Standard deviation [°C]
PI	60,857	1,959
Predictive	60,89	0,9795

Table 4: Product quantities with different control techniques

Control	Top product [t/day]	Bottom product [t/day]
PI	263,4	2873
Predictive	267,2	2869,2

The top product can be increased by 3.8 t/day because of the better control performance. The top product is more expensive than the bottom product in summer by 27 €/t and in winter by 9 €/t, so in average the top product in the whole year is more expensive than the bottom product by (27 + 9)/2=18 €/t. Assuming availability of predictive control in 90 % of the operating time, the benefit becomes (top product increase) x (price advantage) x (availability) x (365 day/year): 3.8 t/day x 18€/t x 0.9 x 365 day/year = 22470 €/year.

The linear control algorithm is valid only near to the working point for changes of about 10 %. In another working point new model estimation and a new (linear) controller tuning would be required.

The above case study from the industry demonstrates well what can be achieved by the optimization of a cost function with a multivariable predictive control under constraints.

16. References

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